UNIVERSITY OF ALBERTA

Master of Engineering Project

Process Control Study of Coupled Distillation Columns

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Summary

Process control study was done on Dow Chemical's coupled distillation columns at its Fractionator plant. Due to variability in feed composition, the first of the two towers, T3, often experienced process upset and then passed it on to the second tower, T4. This could result in sending off-specification products to the customers.

Step test was applied for each input-output loop of the process to identify the transfer function that could represent the model of the two towers. From the results, it was apparent that the sampling time inherent in the GC analyzer was relatively long in comparison to the process time constants of the transfer functions. To overcome this deficiency, it was suggested to adopt the inferential control technique and to use the most responsive tray temperatures as the secondary variable to infer the tower compositions on a continuous basis. The RGA was also used on the established models to determine the best pairing for the control loops. For both towers, it was found that the reflux should be controlling the overhead composition of the tower and the steam should be paired up with the bottom composition to minimize the undesired loop interaction. The result was also consistent with the loop-pairing in the current plant. However, it was also found that there were some degree of loop interaction within T4 tower.

In the proposed control scheme, feedforward controllers and a de-coupler were implemented to counteract the process disturbance and loop interaction respectively. Due to some limitations in process dynamics, some feedforward controllers were replaced by the steady-state feedforward controllers. The proposed control system was effective in lowering the process variability and reducing the possibility of producing off-spec products. With this improvement, it was also recommended to raise the control setpoints closer to quality limits. This would help to operate the process more efficiently with less energy consumption.

Finally, the proposed control scheme demonstrated that it was able to sustain the performance with the ±20% model mismatch in process gain, time constant, and delay time. Although it would create some minor distraction due to model discrepancy, the feedback control was able to compensate it effectively. The tuning for the PI controllers should also be set to ensure process stability at all time. This would prevent the process from becoming oscillatory when experiencing a sudden change in process conditions.

Summary

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1. Introduction

The Dow Chemical's Fractionator plant at its Fort Saskatchewan site contains a coupled distillation columns, namely the de-propanizer (T3) and the de-butanzier (T4) towers. The T3 tower takes the C3+ (propane and heavier components) natural gas liquid feed either from the upstream process of the plant or from the customers via pipeline shipments. It then separates the hydrocarbon mixture into lighter and heavier streams through its multi-staged tray operation. The distillate or the lighter stream exiting from the overhead of the tower is consisted of mostly the propane, small amount of butane, and trace amount of pentane. The distillate product can be either shipped to the customer through pipeline or be stored in the storage caverns. The condensate or the heavier stream from the tower bottom flows directly to the T4 tower for further purification.

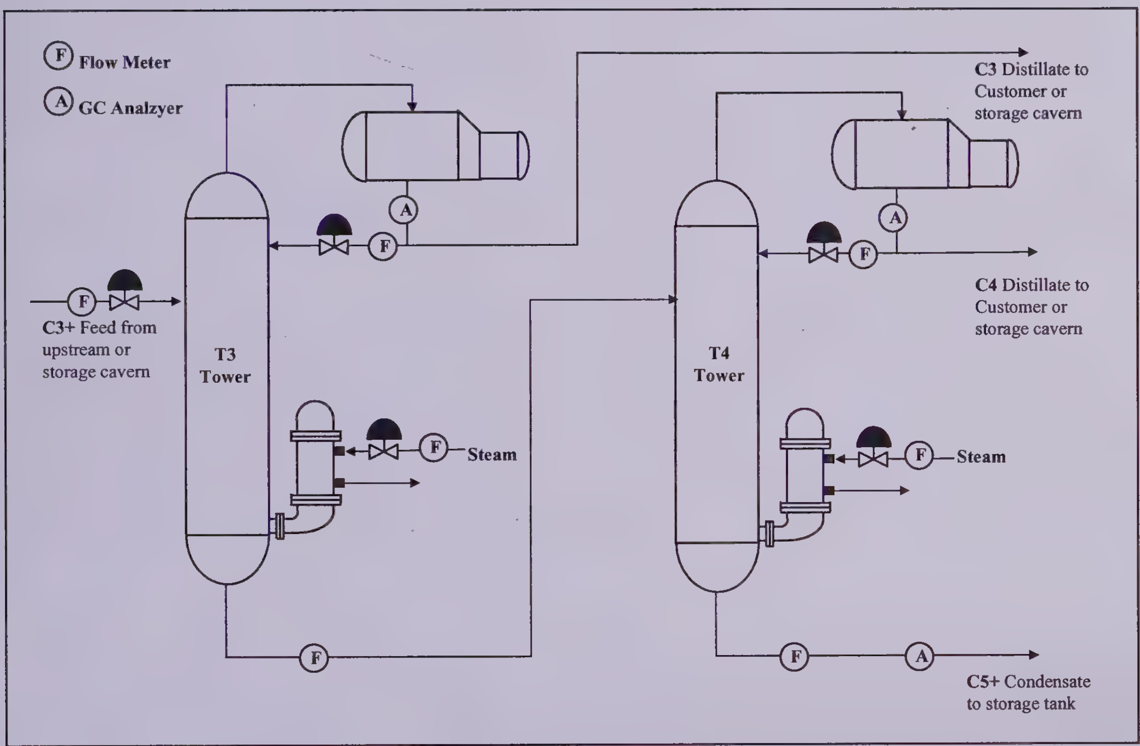

Figure 1. Coupled Distillation Columns Process Schematic

The T4 tower is also a distillation tower with multiple tray design. After taking the feed from its upstream tower, T4 produces the butane-rich distillate from its overhead and the pentane-rich condensate from its bottom stream. Its distillate can be pipeline-transferred or

stored in the storage cavern, while its bottom stream can only be stored in the storage tank waiting for the weekly shipment to the customers. All hydrocarbon products have quality specification requirements. Figure 1 shows a simple schematic of the process.

In the existing control scheme, both towers manipulate the reflux flow and steam flow to achieve the desired overhead and bottom concentration respectively. Stream compositions are monitored by Gas Chromatography (GC) analyzers located in the feed stream to the T3 tower, in the distillate streams for both towers, and in the condensate stream for the T4 tower. These GC's take approximately 12 minutes to automatically analyze the process stream and update the results to the control system. The minor components of the product streams were controlled for better resolution of the process signals. All controllers in the system are simple PI controllers. Table 1 summaries the existing control loop pairing for the process. Note that it was desired to control T3 bottom C3 composition by manipulating T3 steam, but since there was no analyzer available in the bottom of T3 tower, the C3 concentration in T4 overhead stream was used as the alternative controlled variable and paired with the T3 steam flow.

Table 1. Existing Control Loop pairing for the Coupled-Distillation Columns.

Controlled Variable (*CV*)	Abbreviation	Manipulated Variable (*MV*)	Abbreviation
T3 Overhead C4 Concentration	T3_O_C4	T3 Reflux Flow	T3_R_F
T4 Overhead C3 Concentration	T4_O_C3	T3 Steam Flow	T3_S_F
T4 Overhead C5 Concentration	T4_O_C5	T4 Reflux Flow	T4_R_F
T4 Bottom C4 Concentration	T4_B_C4	T4 Steam Flow	T4_S_F

From time to time, feed condition to the T3 tower such as flow or concentration would change suddenly and lead to process upsets. The upset in the first tower could also pass on to the second tower and create more problems for the plant operation. This often results in poor separation of the natural gas liquid and causes the off-spec products to be sent to the customers.

This project studied the process dynamic and control of the coupled distillation columns system and identified any opportunity for improvement. The first part of the analysis included a process (model) identification to estimate the dynamic model of the current plant.

Real plant data was collected and analyzed to derive the best models. The obtained models for the coupled distillation process were then used as the simulation plant in the subsequent analysis:

- Relative Gain Array (RGA) anaysis: This would help to identify whether the existing control loop pairing was adequate. The analysis would also reveal the degree of process interaction between the process variables in the towers.
- Performance study for existing control system: Evaluate the extent of feed disturbance that the existing PI control system can handle without compromising product qualities for both towers.
- Performance study for proposed control system: Propose a new control scheme that can handle the feed disturbance better than the existing scheme. Ideally, it should be able to handle the worse case scenario of process disturbance based on the past historical data.
- Model Mismatch study: Investigate the impact of model mismatch for the proposed control scheme. This is to account for any inaccuracy in the process identification results due to feed composition changes, reboiler fouling, etc.

Finally, due to the confidentiality of process data, all data used in the report was normalized so that any sensitive process information would not be release to other parties.

2. Process Model Identification

The key to the success of this study was built on the simulation model that can represent the dynamic of the plant with reasonable accuracy. To develop such a model, one must select a process identification technique that is best suited for the study.

2.1. *Selection of Process Model Identification Method*

Two common Process Identification approaches were considered initially, the Parametric method and Nonparametric method. The Parametric method uses the plant data and linear regression technique to estimate the unknown parameters in the linear equation that could best represent the plant model. However, this approach would require a large amount of plant test data (i.e. in the order of hundreds or thousands) and some restrictions on the input signals (i.e. minimum signal to noise ratio and linearly independent w.r.t. the output) in order to give a satisfactory result. Applying this to the study, it would require to disturb the running plant by random inputs with significant magnitude and to keep it open-loop for days or even weeks. This would be impractical for an operating distillation process that has specific production rate and strict quality requirements. Another option in the Parametric method was to use the closed-loop plant data to estimate the model parameters with thousands of data already available in the plant historian database. However, the complexity of this technique escalates significantly when dealing with the multi-variable system. If only one input-output pair was analyzed at a time, the interaction from other loops would also get embedded in the model results. Thus, this was not a favorable technique to be employed either.

On the other hand, the Step Response approach in the Nonparametric Method would only require to put the process in open-loop for a relative short period of time. By using an manipulated variable that has adequate input-to-noise-ratio to excite the plant in a step-wise fashion, the process dynamics could be easily observed and interpreted as the transfer functions. Thus, for its simplicity and short duration, this approach was selected to be used for identification of process models.

2.2. Step Response Model

The Step Response identification for each pair of input-output combination was done for the two towers. To study the effect of process disturbance to the plant, the C4 component in feed flow to T3 (T3_C4_Feed) and the T4 overhead C3 concentration (T4_O_C3) were also treated as the additional inputs to T3 and T4 processes respectively. The first disturbance input would help to identify the impact of any feed flow or feed composition changes on the towers. It was calculated as multiplying the feed flow by its C4 concentration. The second disturbance input was also the controlled variable for T3. It would associate the processes of the two towers together. For each input-output pair, the input was manipulated in a step-wise manner while keeping other process inputs as steady as possible. The responses of the outputs were recorded. For the disturbance inputs, particularly the T4 overhead C3 composition, since it was not a manipulated variable that could be regulated easily, this posed some challenges to perform the step test. Thus, its step test was done indirectly by waiting for the process to experience a step-wise (or close to a step-wise) disturbance and obtained the results. In principle, the excitation of the inputs should be as large as possible in order to desensitize the effects of the process noise on the results. The test conditions should also be within the normal operating ranges so that the test data would be meaningful. Nevertheless, during the tests, the excitation magnitudes of these inputs were consolidated with the plant operator and set at the values that would not jeopardize the product quality.

Plot of each step-test is included in Appendix I. Note that the data was normalized to comply with Dow Chemical's process confidentiality requirement and to avoid revealing process sensitive information to other parties. In the normalization, all process inputs and outputs were rescaled such that all variables were bounded between 5 and -5. For inputs, the upper and lower limits would correspond to their actual process constraint, while the limits of the outputs would refer to the product specifications of the towers. The time unit was based on one minute while the discrete composition data was provided by the Gas Chromatography (GC). It took 12 minutes for the analyzer to update the subsequent sample result.

2.3. *Process Model Results*

From each of the step response result, a transfer function can be derived based on the different characteristics of the curve. For simplification, each model was approximated either by a first order or a second order transfer function. Table 2 summaries the model results.

Table 2 Process Model Identification Results

T3	T3_O_C4	T4_O_C3
T3 Reflux (T3_R_F)	$\frac{-4.149}{13s+1}e^{(-30s)}$	$\frac{2.678}{281.2s^2+34s+1}e^{(-100s)}$
T3 Steam (T3_S_F)	-	$\frac{-1.488}{21s+1}e^{(-30s)}$
T3_C4_Feed (disturbance)	$\frac{0.2}{6s+1}e^{(-20s)}$	$\frac{0.52}{23s+1}e^{(-20s)}$
T4	**T4 Ovhd C5**	**T4 Btm C4**
T4 Reflux (T4_R_F)	$\frac{-4.76}{18s+1}e^{(-20s)}$	$\frac{3.288}{30s+1}e^{(-40s)}$
T4 Steam (T4_S_F)	$\frac{0.798}{83.3s^2+21.7s+1}e^{(-28s)}$	$\frac{-2.531}{14s+1}e^{(-30s)}$
T4_O_C3 (disturbance)	$0.375e^{(-10s)}$	$0.381e^{(-40s)}$

The step response of T3_O_C4 did not show any significant response subject to the T3 steam change during the step test. This suggested that the steam flow in T3 did not have an impact on the T3 overhead composition.

Another discovery was made that the sampling time of the process (i.e. analyzing time for the GC) was quite large in comparison to the process time constants for all transfer functions. Relying on the feedback of the GC's to control the tower composition may face some challenges due to the inappropriate sampling time that took too long to update.

3. Current Control System Evaluation

The current control system and process models of the plant were studied and simulated to identify the opportunities for control improvements.

3.1. Relative Gain Array Analysis

To design a control system for a multivariable system such as a distillation tower, it is beneficial to identify and pair the outputs with their most effective inputs in the system. A well-paired control loop system can reduce the undesired interaction of process variables and increase the performance of the control loops. The Relative Gain Array (RGA) analysis can be used to determine the most effective pairing of the control loops by evaluating the steady-state gains and interactions between the input and output pairs. Following the scheme as described by Ogunnaike et al. (1994), the RGA, Λ_{T3} and Λ_{T4}, for the 2 x 2 system of each tower can be determined as:

$$\Lambda_{T3}\begin{bmatrix} T3\ Ovhd\ \ C4 \\ T4\ Ovhd\ C3 \end{bmatrix} = \begin{bmatrix} 1 & 0 \\ 0 & 1 \end{bmatrix}\begin{bmatrix} reflux \\ steam \end{bmatrix}$$

$$\Lambda_{T4}\begin{bmatrix} T4\ Ovhd\ \ C5 \\ T4\ Btm\ C4 \end{bmatrix} = \begin{bmatrix} 1.278 & -0.278 \\ -0.278 & 1.278 \end{bmatrix}\begin{bmatrix} reflux \\ steam \end{bmatrix}$$

A detailed calculation for Λ_{T3} and Λ_{T4} is also provided in the Appendix II. In RGA analysis, the ideal pairing would be the input-output pair with the matrix element of 1, indicating that the open-loop gain between the output and input is identical to the closed-loop gain. For T3, since the steam flow showed negligible effect on the overhead composition, this yielded the ideal RGA matrix and implied that the overhead C4 should be paired with the reflux while the T4 overhead C3 should be controlled by T3 steam. On the other hand, the diagonal elements in the T4's RGA matrix have the values of 1.278. This suggested that the control loops interacted and the retaliatory effect from the other loops acted in opposition to the main effect of reflux (steam) on T4 overhead C5 (T4 bottom C4). But, since the diagonal elements were only slightly above 1, the recommended pairing of T4 would be similar to T3, that is to have

the reflux controlling the overhead and the steam controlling the bottom composition.

In addition to the RGA, the dynamics of each loop should also be considering when choosing the appropriate loop pairing in order to avoid structural instability as well. Since the RGA is based only on the process gain, it can easily happen that the best RGA pairing resulted in very sluggish closed-loop response because of long time delays, significant inverse response, or large time constants. In this situation, choosing a pairing with inferior RGA properties but without sluggish dynamics could greatly improve the control system performance. In Table 2, the diagonal transfer functions for T3 and T4 had shorter delay time and smaller time constants, and thus had faster process responses than their opposite loops. This also agreed with the results from RGA for the best loop pairing.

The results from above are consistent with the current control system in the plant by utilizing the reflux to maintain the tower overhead purity and the steam to manage the bottom composition for both towers. However, it did indicate that there was some degree of interaction between steam and reflux on both the overhead and bottom qualities in T4 tower. Furthermore, the transfer functions of the disturbance implied that the feed to T3 would cause significant disturbance to the product quality of the towers. The distillate composition from T3 could also impact the T4 as well. These all identified the existence of opportunities for the current control system in the plant.

3.2. Sampling Time Constraint

Although the current control system in the plant has the capability of providing regulatory control action every second, the analyzers monitoring the tower concentration could only provide update every 12 minutes. As a Rule-of-Thumb in the control applications, the sampling times should be normally set in the range of 0.1 ~ 0.2 of the process time constant in order to monitor the process response appropriately. In this process, the slow analyzing time imposed by the GC's does not meet this requirement. Excess long sampling time would impair the controller response of the system and make the system vulnerable to abrupt disturbance change. But, since this limitation could not be improved without changes in the analyzer hardware, it was proposed to adopt the inferential control in the system by finding an alternative process measurement that is continuous and yet can provide a good indication of

the tower's operating condition. This inferential control strategy is to infer the composition (known as the "primary variables") based on other process variables such as the most responsive tower tray temperature (known as the "secondary variables"). It has gained popularity in the industry today in order to deal with the large time delay and high failure ratio associated with the analyzers.

With the availability of several temperature measurements within each tower, it is suggested to use the temperature of the most responsive tray, the tray immediately above, and the tray below to calculate an average temperature as the secondary control variable. There are two advantages by utilizing multiple trays. First, it can monitor the tower process better by checking the average response of the trays. Since the operation of a distillation tower is generally characterized by the temperature profile of the trays, using the average tray temperature would represent the tower operation better than a single point temperature. Secondly, the reliability of the control system would also be enhanced by employing multiple tray temperatures. A failure in any single temperature transmitter would not cripple the control system.

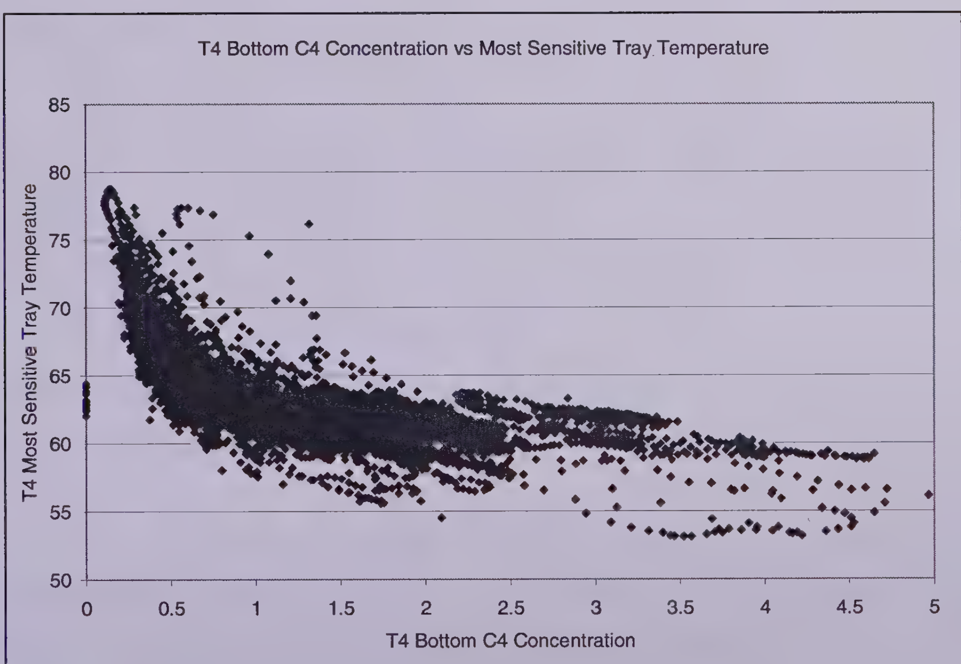

Figure 2. T4 Most Responsive Tray Temperature versus Bottom Composition (Normalized)

Figure 2 illustrates the relationship between the primary variable and secondary variable in

the inferential control scheme. For implementation, one would consider to describe the relationship either by a polynomial equation or a logarithm function. However, from the scattered data deviating from the main curve, this was also suggest that the established function could be drifted by process factors such as changes in feed or operating conditions. Thus, a continuous bias update should be implemented in the control scheme to account for this.

Since the main focus in this study was to examine the relationship between the product compositions and the input variables such as steam and reflux, the tower temperature that served as the secondary variable to the composition was excluded in the analysis. Nevertheless, for the rest of this study, it would be assumed that the inferential control scheme was implemented such that continuous indications of tower composition were available. Therefore, the continuous models of the plant would be used in the simulation. The implication and effects of potential error in the inferential function and bias would be discussed as part of the model mismatch in section 5 of this report.

3.3. Simulation on Existing Control System

The plant dynamics were simulated in the Matlab Simulink program by using the determined plant models and Proportional-Integral (PI) controllers to match the existing control system. The magnitudes of the setpoint changes were to match the setpoints in the current plant. The size for the step disturbance from the T3 C4 feed was determined based on the maximum feed increase that the plant experienced during the two months of the project period. This was taken as the worst case scenario for the feed change. A schematic of the plant simulation in Simulink is shown in Figure 3.

To simulate the process constraints of the manipulated variables, each process was bounded between the established range of -5 and 5. This would prevent the steam or reflux to go beyond any point that is not realistic physically in the actual process.

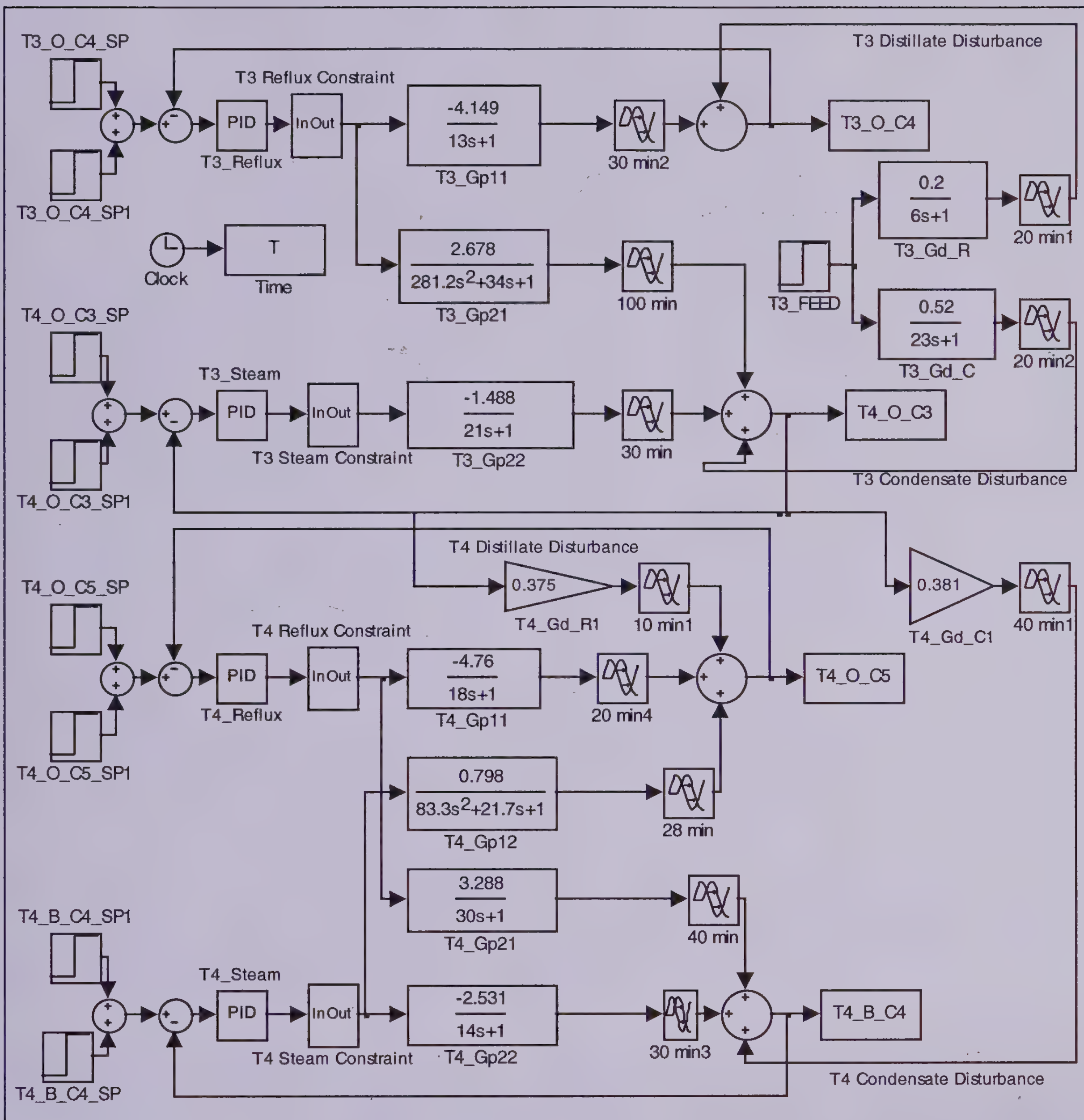

Figure 3. Plant Schematic in Simulink

The Internal Model Control (IMC) tuning algorithm, commonly used in Dow's control applications, was employed to obtain the initial tuning parameters, and then followed by fine tuning to obtain the more optimal settings. Since some of the control loops in the system experienced interaction from other loops and most transfer functions had significant process deadtime (i.e. deadtime longer than the process time constant), a more conservative setting was used to start with the tuning process. A summary of tuning parameters used in the simulation is shown below.

$$K_C = \frac{\tau_P}{\tau_{CL} + \tau_d} \cdot \frac{1}{K_P}; \ \tau_I = \tau_P; \text{ for conservative tuning } \tau_{CL} = \tau_P / 2$$

Table 3 Tuning Parameters of the Control-Loops

	K_p	τ_p	τ_d	K_c	τ_I
(T3 Ovhd C4, T3 Reflux) *IMC Tuning Rule*	-4.15	13	30	13/(13/2+30)/-4.15= -0.09	13
(T4 Ovhd C3, T3 Steam) *IMC Tuning Rule*	-1.49	21	30	21/(21/2+30)/-1.49= -0.22	21
(T4 Ovhd C5, T4 Reflux) *IMC Tuning Rule*	-4.76	18	20	18/(18/2+20)/-4.76= -0.09	18
(T4 Btm C4, T4 Steam) *IMC Tuning Rule*	-2.53	14	30	14/(14/2+30)/-2.53= -0.14	14
(T3 Ovhd C4, T3 Reflux) *after fine tuning*	-4.15	13	30	-0.13	22.5
(T4 Ovhd C3, T3 Steam) *after fine tuning*	-1.49	21	30	-0.57	43.1
(T4 Ovhd C5, T4 Reflux) *after fine tuning*	-4.76	18	20	-0.16	24
(T4 Btm C4, T4 Steam) *after fine tuning*	-2.53	14	30	-0.20	25.2

The simulation first introduced a sequence of setpoint changes for each of the control loop and followed by a T3 feed disturbance as demonstrated in Figure 4. This would allow us to observe the impact of each loop on the entire process.

After running the simulation, the process inputs or manipulated variable (MV) and the process output or controlled variable (CV) were plotted for each of the loop. Figure 5 shows the CV action and setpoint on the upper section and the MV action in the lower section. One can see that the feed disturbance can lead to significant process upset for T3 overhead product with its concentration peaking toward the specification limit of 5 at time 3500. Even though the composition still remained within the specification, any improvement that can be made to reduce the process variability will be beneficial to the operation.

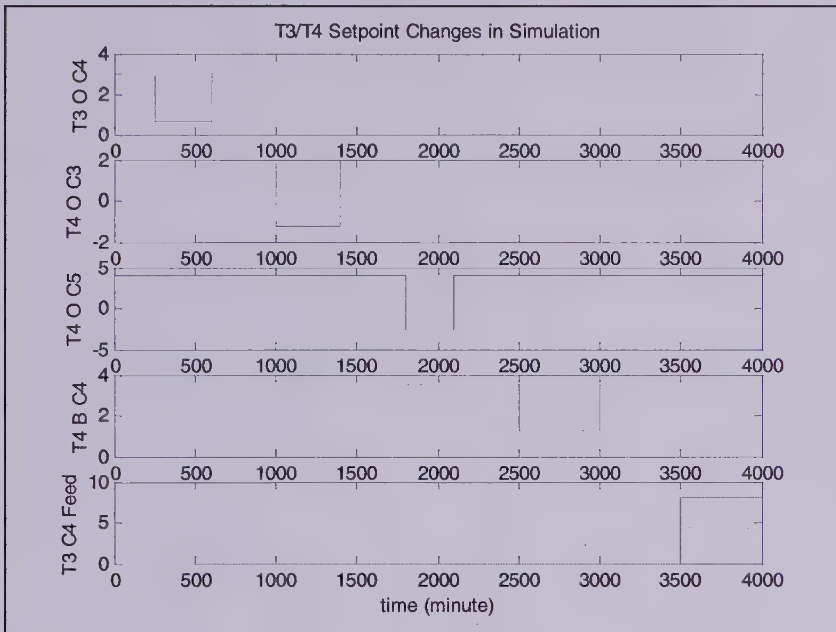

Figure 4. Sequence of process and disturbance setpoint changes in simulation

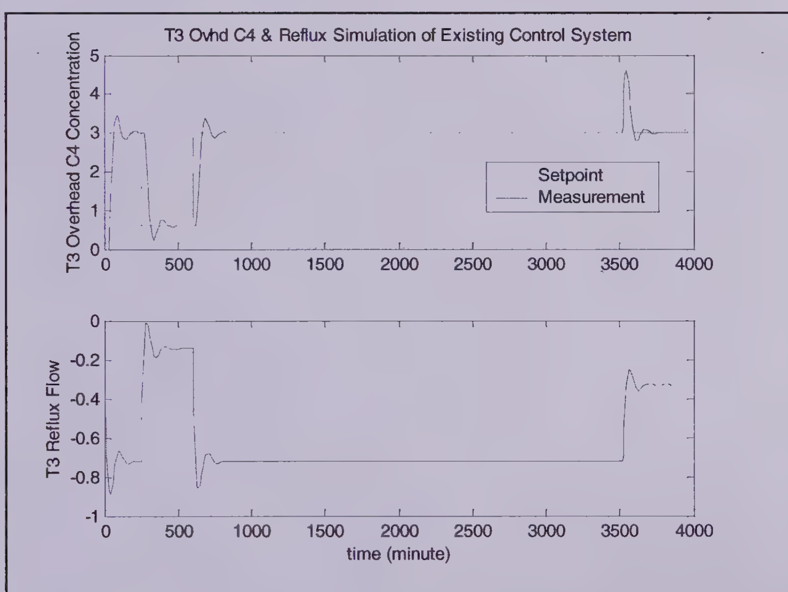

Figure 5. Simulation of existing T3 Overhead C4 and Reflux control loop

In Figure 6, T3 bottom product, denoted by T4 overhead C3, experienced some degree of interference from T3 Reflux in time interval between 0 and 800. However, in comparison to the impact caused by the T3 Feed disturbance at time 3500, this interference was insignificant. Some control improvement should be made to this control loop in order to prevent T3 bottom product from going off-specification (i.e. concentration great than 5) due to large T3 feed increase.

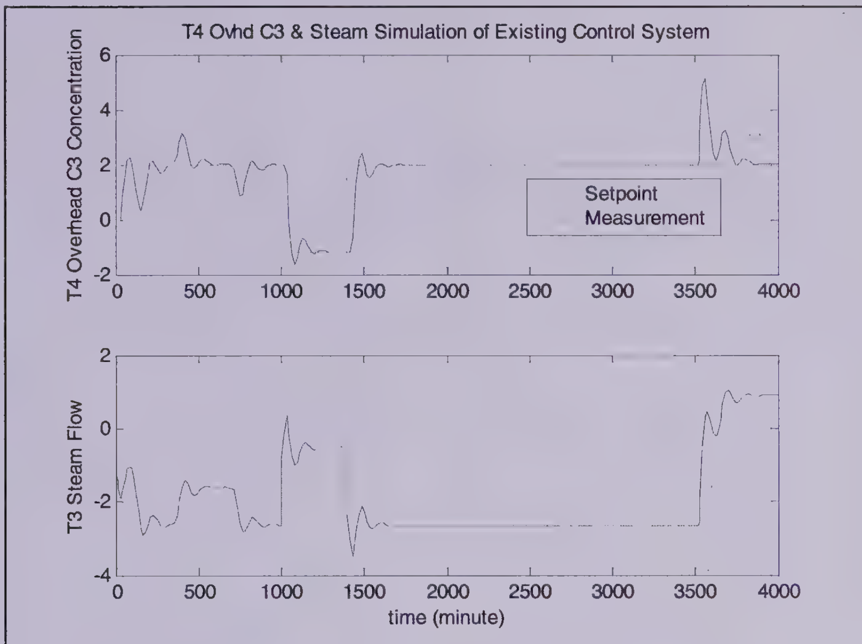

Figure 6. Simulation of existing T4 Overhead C3 and Steam control loop

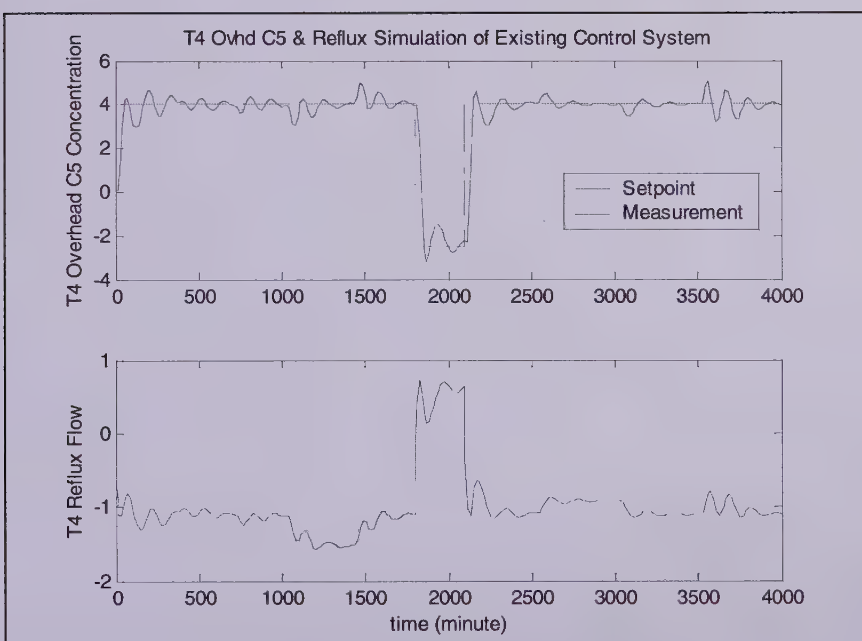

Figure 7. Simulation of existing T4 Overhead C5 and Reflux control loop

The T4 overhead C5 composition, controlled by T4 Reflux flow, did not show any dominant variability in Figure 7 as seen in the previous loops. Nevertheless, its minor distractions caused by the T4 overhead C3 composition variability could be further reduced if the T4 overhead C3/T3 Steam control loop performance could be improved. No immediate attention

is needed to improve the control performance in this loop.

The steam–controlled T4 bottom C4 composition in Figure 8 revealed that it was under substantial interactions from other process variables including T4 Reflux and T4 overhead C3. Specifically, a sizable change in reflux flow at time 1800 could easily drive the bottom product quality exceeding its specification. Without a proper control scheme to effectively reduce these interactions, the process could not be operated efficiently due to the product quality concern.

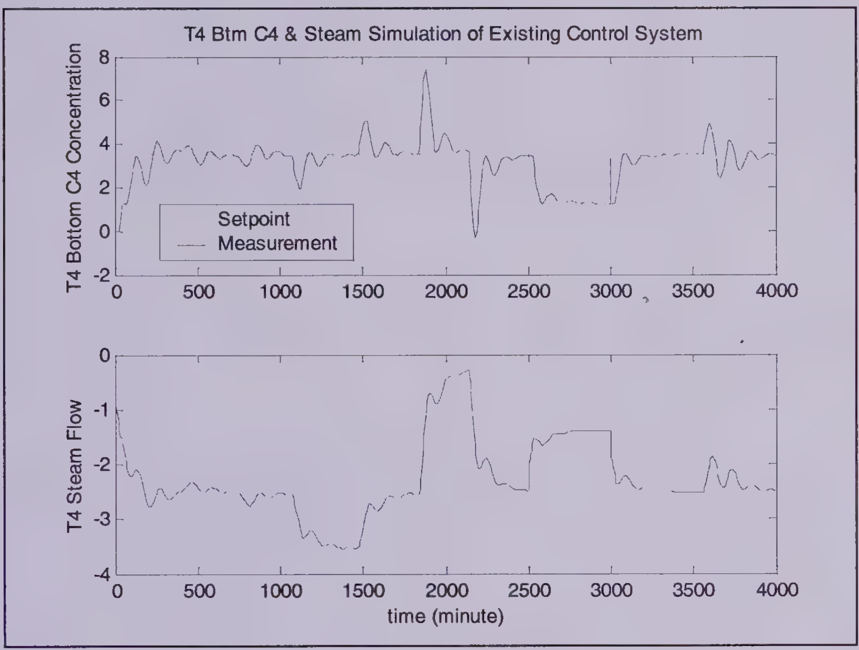

Figure 8. Simulation of existing T4 Bottom C4 and Steam control loop

The implications from Figure 8 is also consistent with the RGA analysis that there were some noticeable interactions between the loops in the T4 tower, and some improvements were needed.

4. Proposed Control System Design

4.1. Decoupling Control Design

With the limited ability of Dow's DCS system to provide complex computation, not all options of the advanced control schemes could be considered without additional hardware cost. Thus, it was recommended to adopt the feedforward and decoupling schemes as simple and effective techniques to compensate the disturbance and undesired loop interaction in the process. The de-coupler was essentially a feedforward controller and was designed by taking the negative disturbance transfer function (Gd) and dividing by the process transfer function (Gp):

$$Gff = -\frac{Gd}{Gp}$$ where Gff = Feedforward controller or De-coupler.

This would allow the manipulated variable to respond appropriately and early in order to offset the process impact from the disturbance. One restriction in the design of an effective feedforward controller was that the response of the disturbance should be slower than the response of the manipulated variable. In other words, if the disturbance would affect the process faster than the manipulated variable, the manipulated would not be able to counteract the disturbance impact before it perturbed the process. Therefore, a disturbance transfer function should have its deadtime and model order larger than the process transfer function. In this study, there were a few instances that did not meet this requirement. They were the T3 feed disturbance to the T3 overhead and bottom compositions, and the T4 overhead C3 disturbance to the T4 overhead C5 and bottom C4 compositions. In these cases, steady-state decouplers were designed and implemented by using the ratio of the steady-state gain rather than using the whole transfer functions. Appendix III provides a sample calculation for the design of these feedforward controllers and de-coupler. Table 3 below summarizes the results. Figure 9 shows the schematic of proposed control system in Simulink.

Table 4 Summary of Feedforward and De-coupler Design

	T3_O_C4	T4_O_C3	T4_B_C4
T3_C4_FEED	G_{ff31}=0.048	G_{ff32}=0.349	-
T4_R_F	-	-	$G_{ff42}=\frac{18.19s+1.3}{30s+1}e^{-10s}$
T4_O_C3	-	-	$G_{ff41}=0.15e^{(-10s)}$

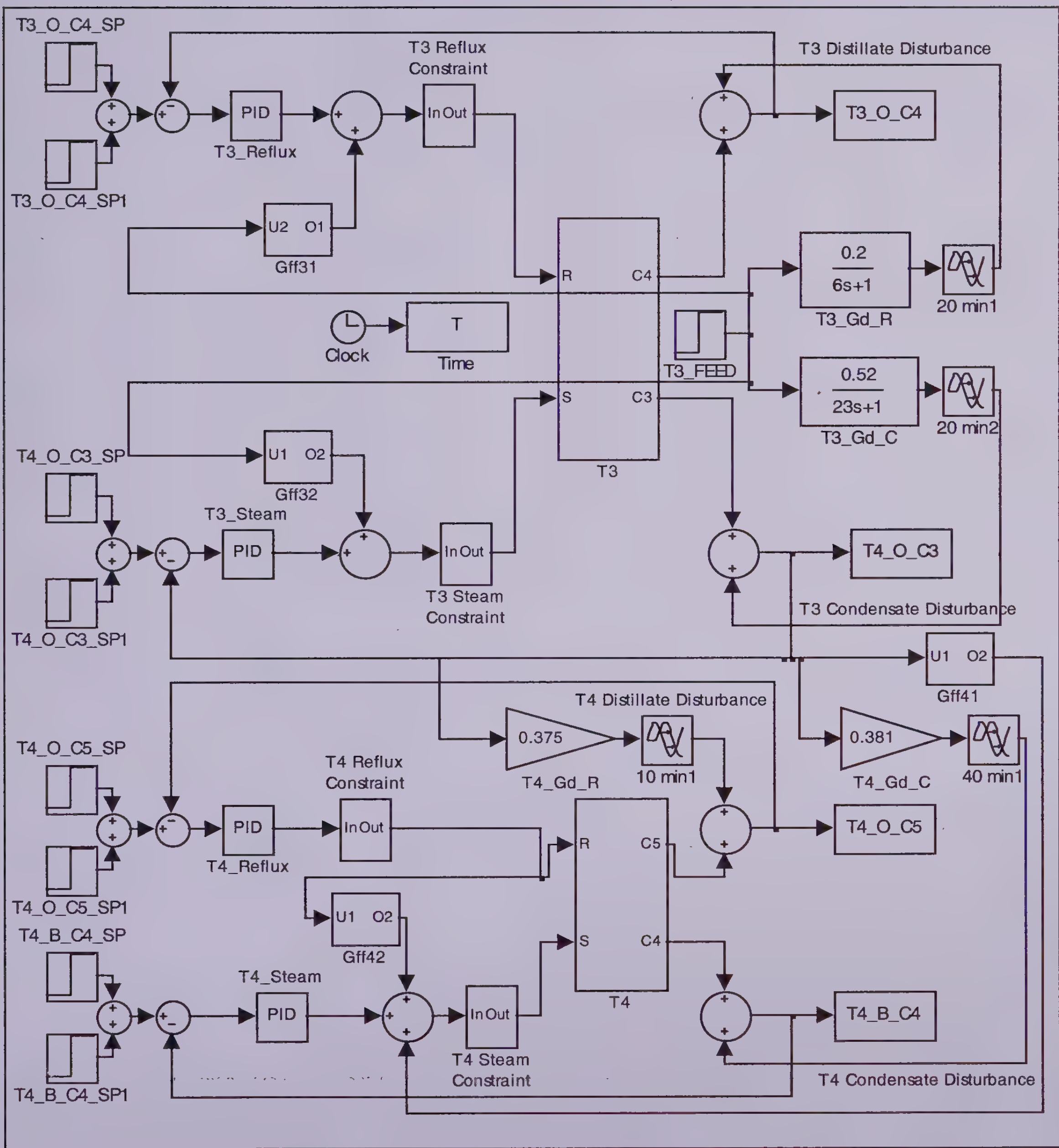

Figure 9. Proposed Control System in Simulink

4.2. Simulation on proposed control system

The proposed de-couplers were implemented and tested in the plant simulation. With the new control system in place, the performance of the T3 overhead C4 and Reflux loop showed some improvements as seen in Figure 10. Although the process spike initiated by T3_C4_Feed disturbance was lowered with the aid of the steady-state feedforward control, it was not able to completely eliminate the disturbance impact due to the longer deadtime of the manipulated variable than the disturbance.

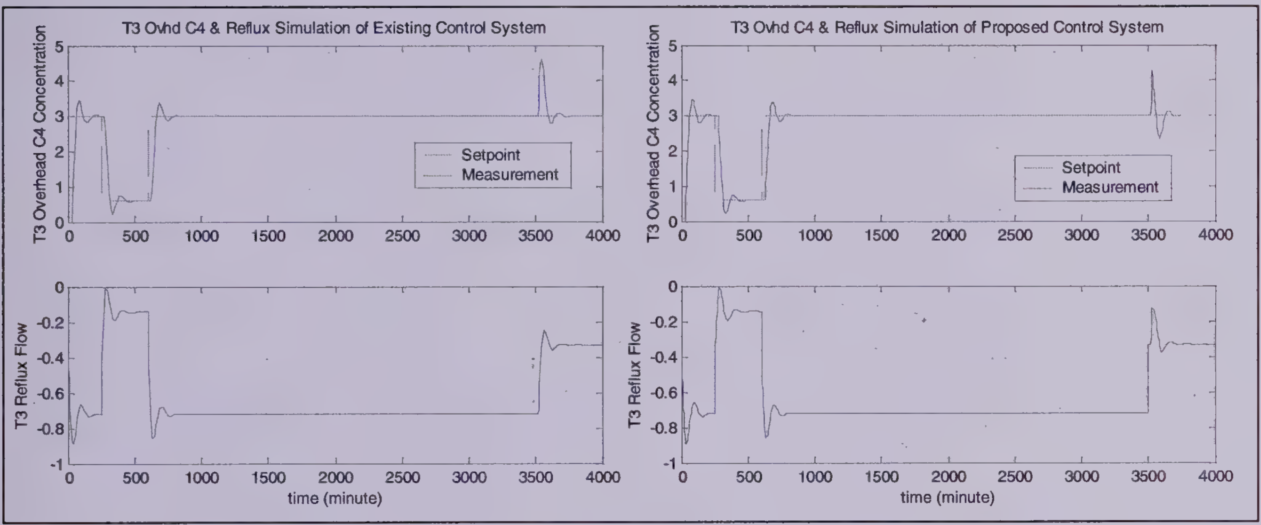

Figure 10. Existing vs Proposed T3 Overhead C4 and Reflux control loop

Similar to T3 overhead control improvement, the T4 overhead C3 and T3 Steam loop also observed a reduction in process variability under the influence of the steady-state feedforward action. As shown in Figure 11, the magnitude of the peak at time 3500 was lowered from exceeding specification limit of 5 (left) to below 4 (right).

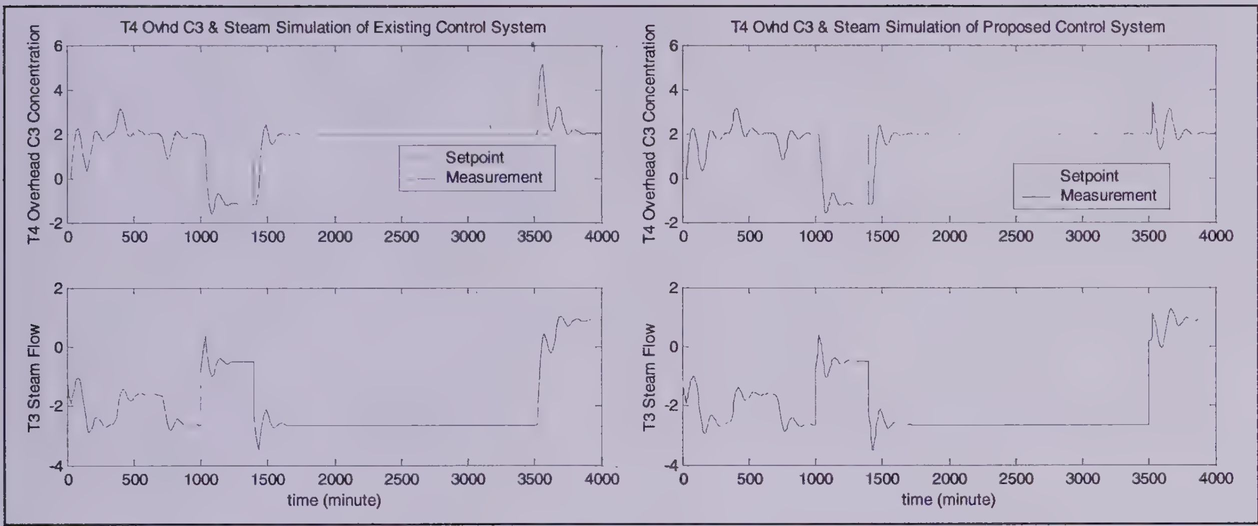

Figure 11. Existing vs Proposed T4 Overhead C3 and Steam control loop

While no control improvement was made to the T4 Overhead C5 and Reflux loop, it still experienced much reduced process variability with the new control system in place. It benefited from the improved control performance in the T4 Overhead C3 and Steam loop and resulted in smaller process disturbances to the process. Figure 12 shows the results.

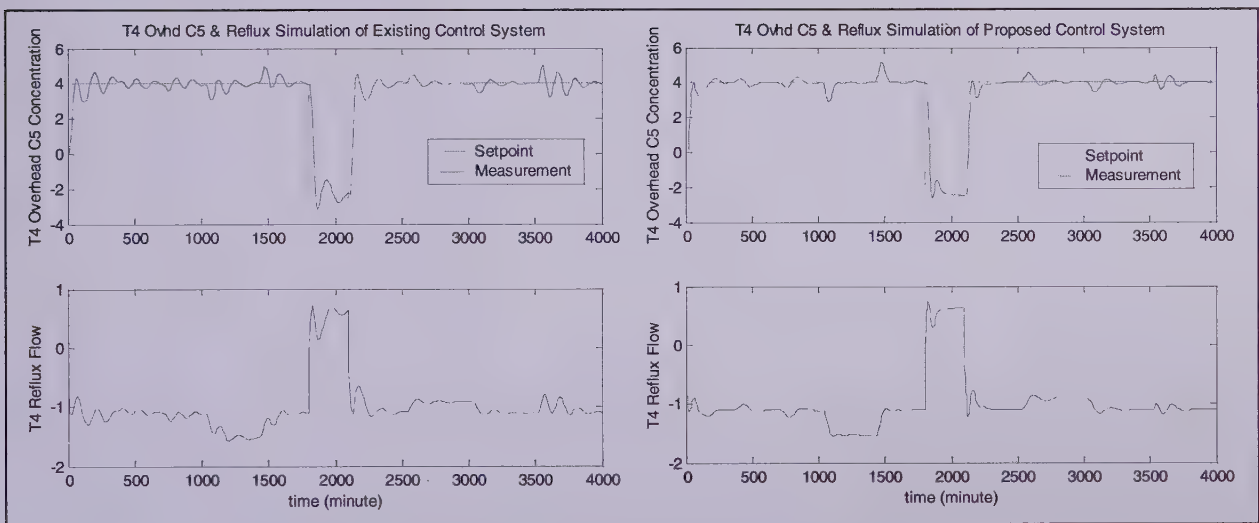

Figure 12. Existing vs Proposed T4 Overhead C5 and Reflux control loop

The performance for the T4 Bottom C4 and Steam loop in Figure 13 demonstrated the most significant improvement among all the loops. First of all, the loop interaction from T4 Reflux was completed eliminated at time 1800 with the de-coupler in place. This gave the best result

of a de-coupler design controller one could achieve. The steady-state feedforward controllers to compensate the T4 Overhead C3 disturbance also had some effectiveness on reducing process variability at time 3600.

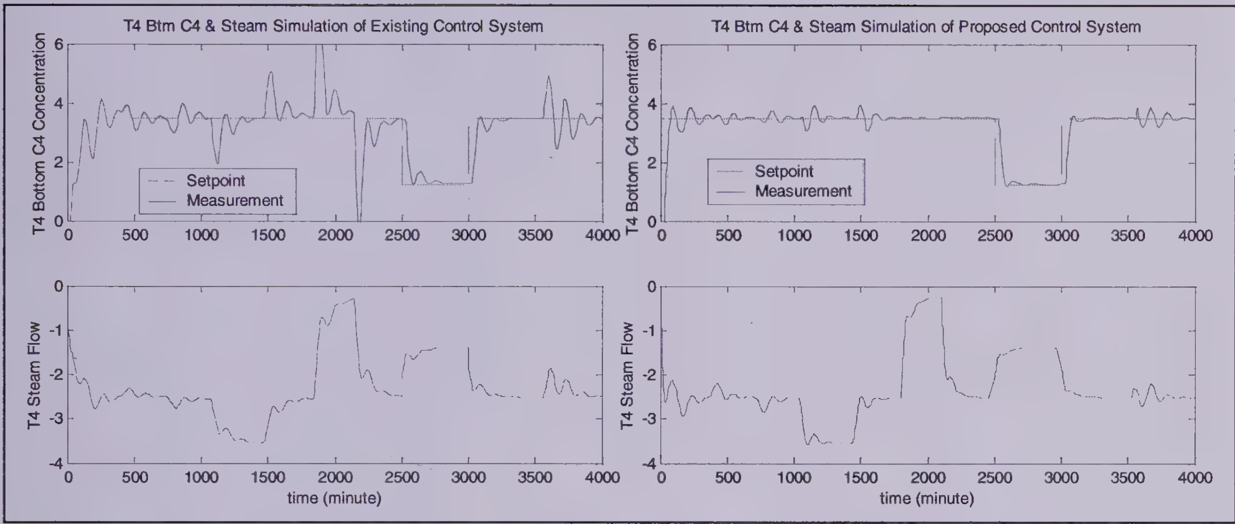

Figure 13. Existing vs Proposed T4 Bottom C4 and Steam control loop

5. Model Mismatch Analysis

For a model-based control system, the results greatly depend on the accuracy of the established plant model. The cause of discrepancy between the actual plant and the model could be contributed by several factors. First, improper techniques in the Process Identification practice could easily generate faulty models. This includes insufficient input-to-noise ratio used to excite the plant in the step test, failing to acknowledge other disturbance that might affect the process during the test, selecting incorrect models to fit the data, etc. Second, a time-varying plant could also depart the plant model from its true dynamic. Any season-varying processes (i.e. cooling water temperature), gradually fouled heat exchangers, and processes with time-varying delay were some of the examples in this category. Third, a change in the state of the process different from the one when the process identification was performed could alter the model quality as well. The changes in feed and operating conditions (i.e. pressure or level of the tower) were under this category. The first type of model mismatch can be minimized by using proper techniques in the Process Identification step. For the second and third types of model mismatch could rely on robustness of the control system to handle the model discrepancy. To evaluate the effect of model mismatch on the proposed control system, the plant model in the simulation were adjusted while keeping the control system and the tuning parameters unchanged. The adjustment was done one at a time on the process gain, process time constant, or delay time for all transfer functions with a magnitude of plus or minus 20%. The ±20% mismatch in process gain was based on the variability of reboiler duty in operation for the last two years. Same degree of mismatch was also applied to process time constant and delay time mismatches, but unlike the process gain mismatch, these were chosen intuitively. Based on the knowledge of the process, a 20% mismatch in process time or deadtime is unlikely but possible, and thus is treated as the worse case scenario in this study. To determine the robustness of the new control system, one can observe if all concentrations would remain within the quality limit of 5 in the presence of model mismatch. This would imply that the system is capable of producing in-specification products despite of model mismatch.

5.1. Process Gain Mismatch

The gain mismatch of the plant was simulated by increasing and decreasing the gain of all transfer functions in the system by 20%. By doing this, one can observe the effect of under-estimating and over-estimating the process gain for each loop. Figures 14 - 17 show the simulation results for each of the control loop.

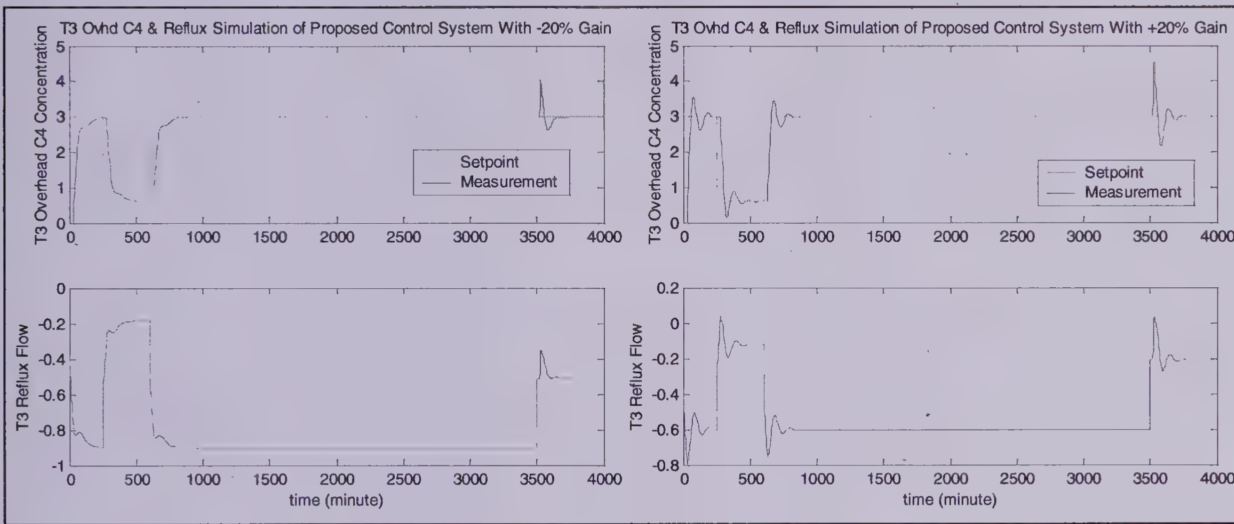

Figure 14. Comparison of T3 Ovhd C4/T3 Reflux Process Gain Mismatch

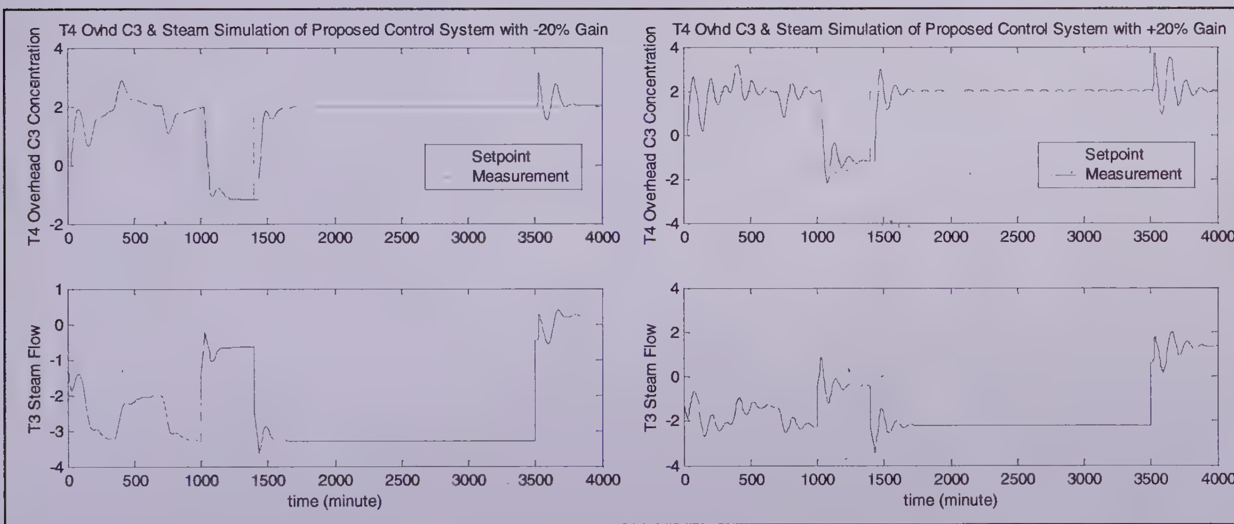

Figure 15. Comparison of T4 Ovhd C3/T3 Steam Process Gain Mismatch

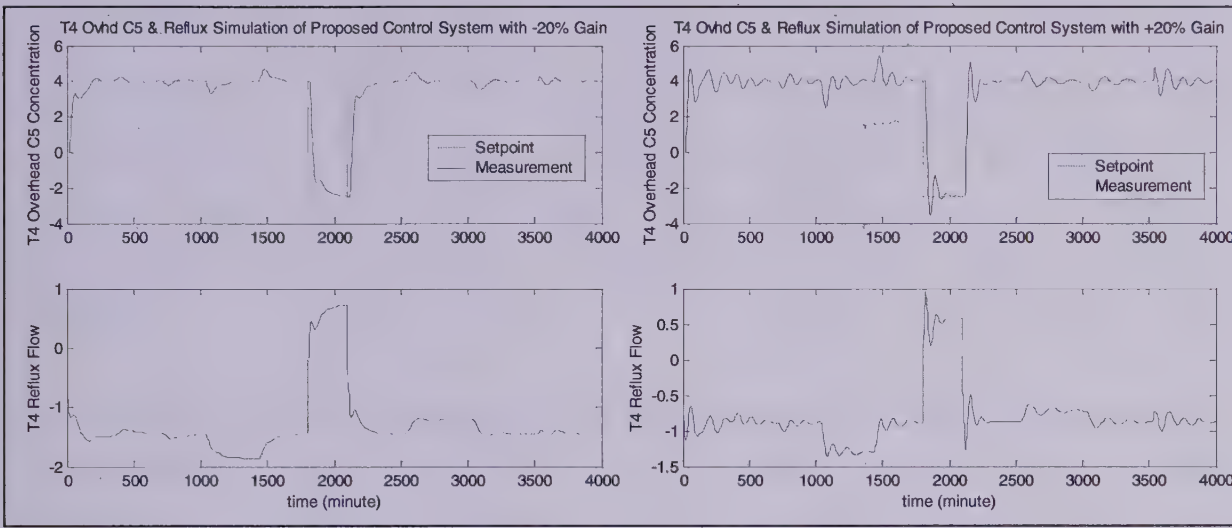

Figure 16. Comparison of T4 Ovhd C5/T4 Reflux Process Gain Mismatch

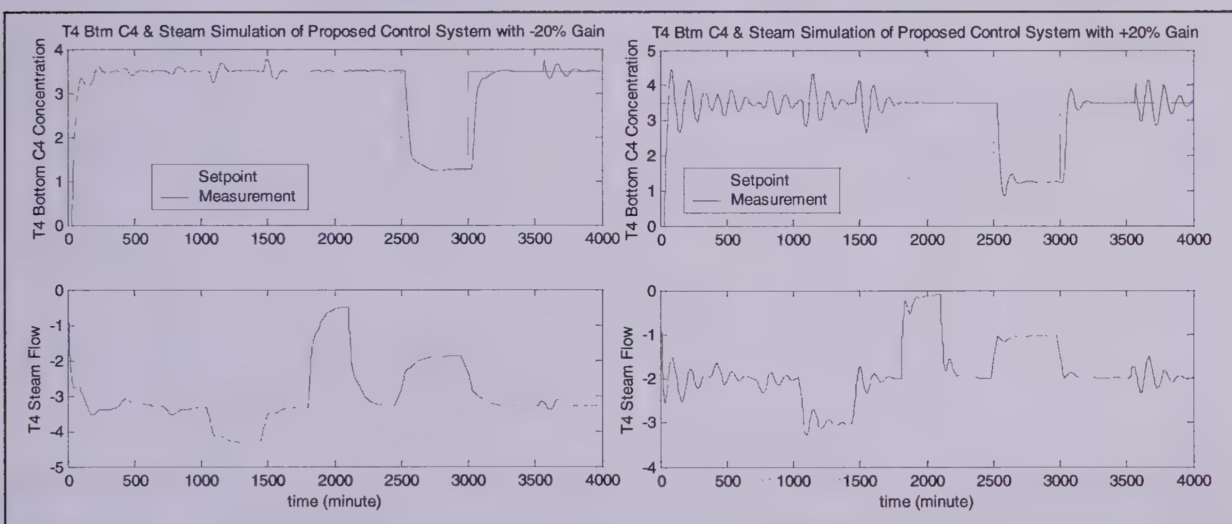

Figure 17. Comparison of T4 Btm C4/T4 Steam Process Gain Mismatch

All previous four graphs indicate that the -20% gains case performed better than the +20% gain cases with less variability. This is because the gains for all of the process and disturbance models were magnified by 20% in the +20% cases. Thus, the same magnitude of disturbance introducing to the system would produce a more significant impact. As the gain varied, the action of the pre-tuned PI controllers also demonstrated some degree of variability. It was found that the controllers tended to be more sluggish for the -20% cases, and more aggressive for the 20% cases. This implied that one set of tuning parameter might not be able

to provide the optimal control performance as the process gain varied. Nevertheless, in Figure 17 the gain mismatch did not affect the performance of the de-coupler significantly as shown in time interval of 1800 to 2100 when there was significant change in reflux flow. In general, all loops except T4 Ovhd C5/T4 Reflux in Figure 16 were able to maintain the concentration within specification limit. For the exception, one might have to either improve the loop tuning or lower the control setpoint (i.e. from 4 to 3.5) to meet the quality requirement.

In the T3/T4 tower process, the main contributors for the process gain changes were the variation in feed conditions and the fouling of the reboilers over time. Since the feed conditions such as its temperature and composition would change quite frequently depending on the feed source, it was impractical to adjust the tuning all the time. Thus, the controller tuning should be set at the conservative level where the process was stable under all circumstances. Even though this would penalize the controller performance from time to time, the stability of the system would be maintained. When implementing the inferential control scheme, the inaccuracy in the equation that describes the relationship between the primary and secondary variable could also lead to process gain discrepancy. In this case, the control scheme should incorporate an adjustable bias term to account for the difference regularly.

5.2. Process Time Mismatch

Similar to the process gain mismatch analysis, a 20% deviation was also applied to the process time constants of the transfer functions to examine the effect of the mismatch. Figures 18 - 21 show the simulation results for each of the control loop.

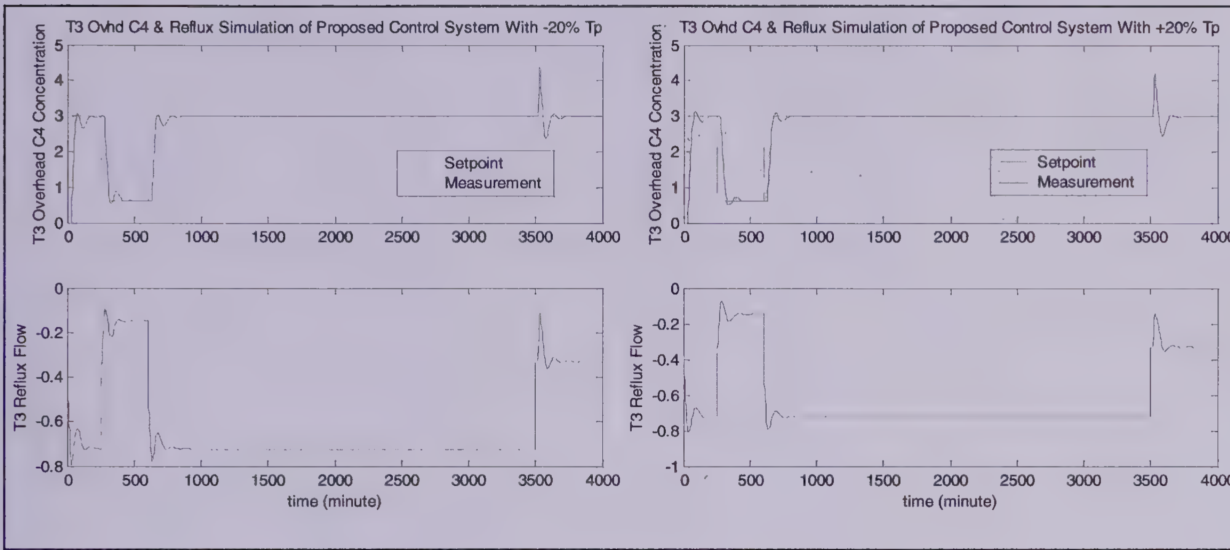

Figure 18. Comparison of T3 Ovhd C4/T3 Reflux Process Time Mismatch

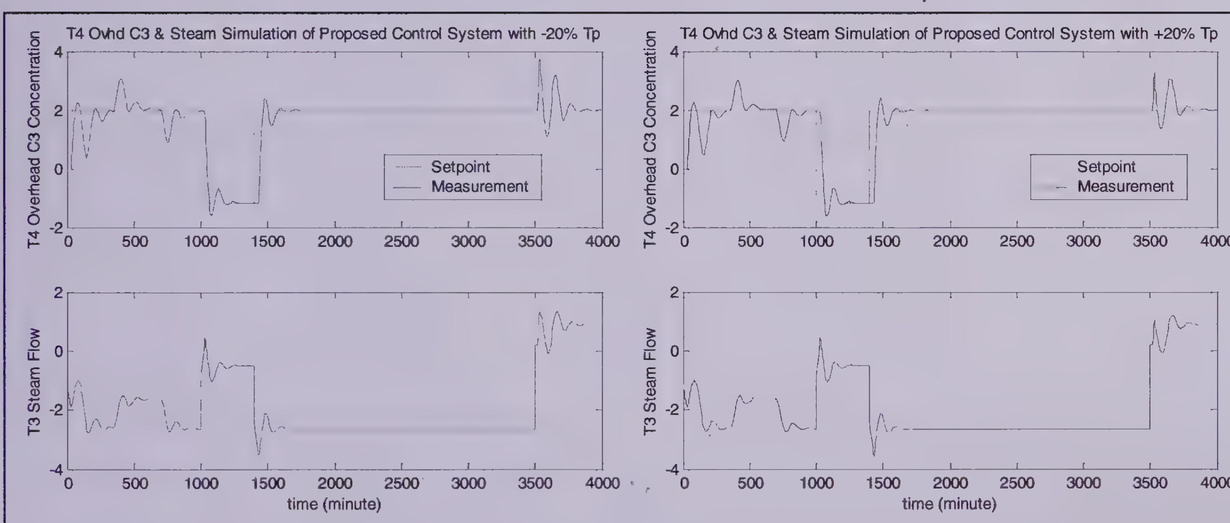

Figure 19. Comparison of T4 Ovhd C3/T3 Steam Process Time Mismatch

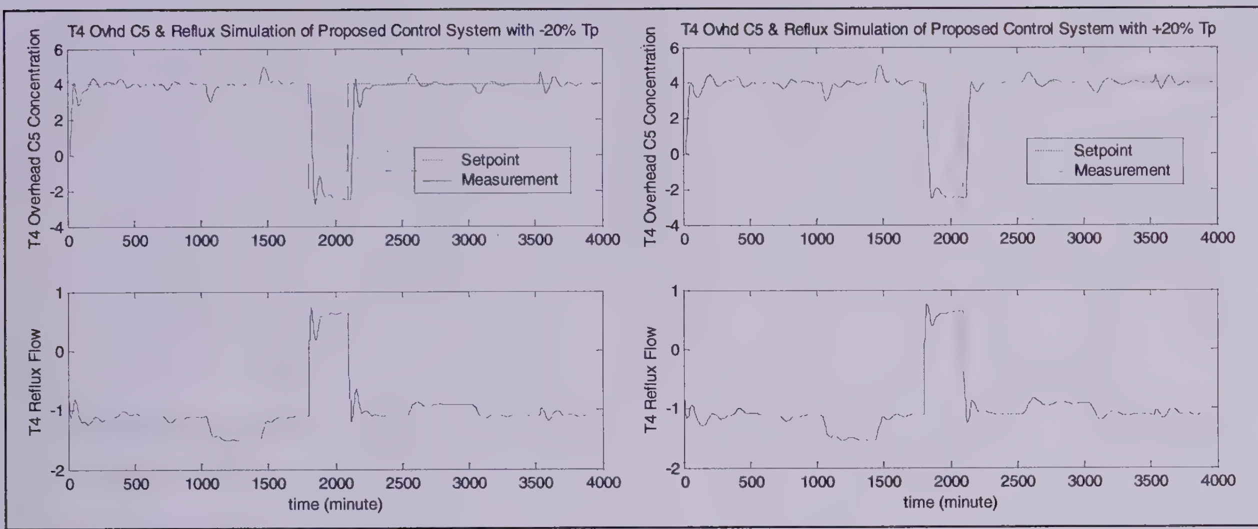

Figure 20. Comparison of T4 Ovhd C5/T4 Reflux Process Time Mismatch

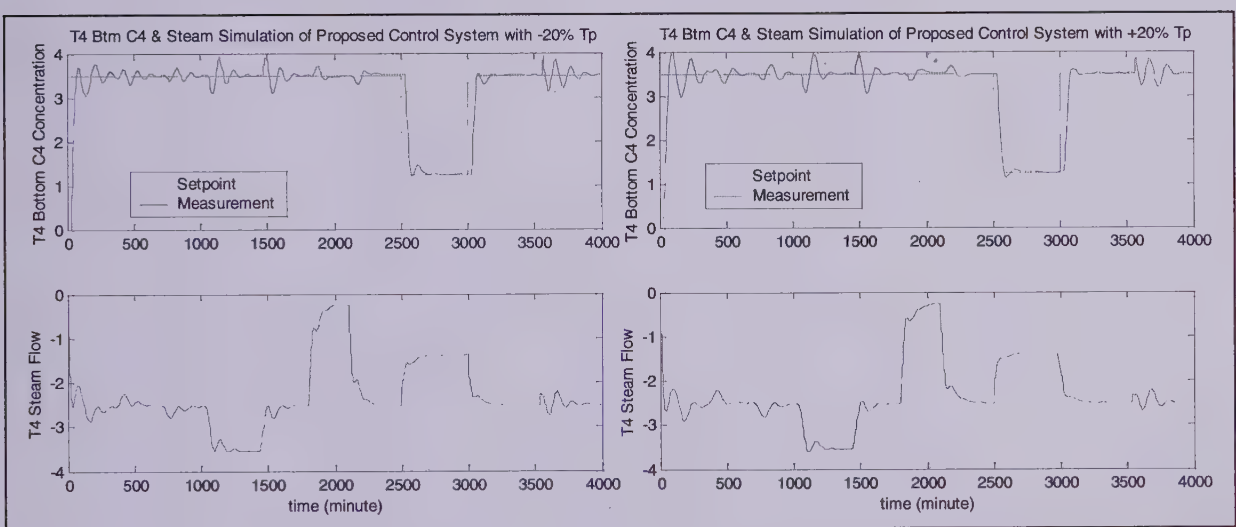

Figure 21. Comparison of T4 Btm C4/T4 Steam Process Time Mismatch

From the graphs, it could be seen that the control performance were better for higher process time cases than the lower process time cases. This was expected as a larger process time constant represented a slower process and thus allowed for the controller to respond more effectively. However, the impact of the process time mismatch was not found as significant as the process gain mismatch did. All control loops were able to keep the concentrations below the specification limit.

The process time mismatch could be easily caused during the curve fitting in the model identification process. Due to the large sampling time with respect to the process time, the time constants for several transfer functions were obtained based on 4 to 5 data points. Statistically, insufficient amount of data points would reduce the accuracy in the estimation of the process time constants. Another potential for process time mismatch was the changed in the operating condition. Changes in the tower pressure or the bottom level would influence the effectiveness of the reflux and steam respectively. For example, if the pressure was increased in the tower for any operational reasons, the vapor would become relatively easier to be condensed and thus affected the reflux dynamics in the process. Also, when the level in the bottom of the tower was reduced, steam would be able to boil smaller volume of the liquid faster, and thus reduced the process time constant of the system.

5.3. *Process Delay Time Mismatch*

The delay time mismatch was also simulated by adding or subtracting 20% of the existing delay time in each transfer function. Figures 22 - 25 show the simulation results for each of the control loop.

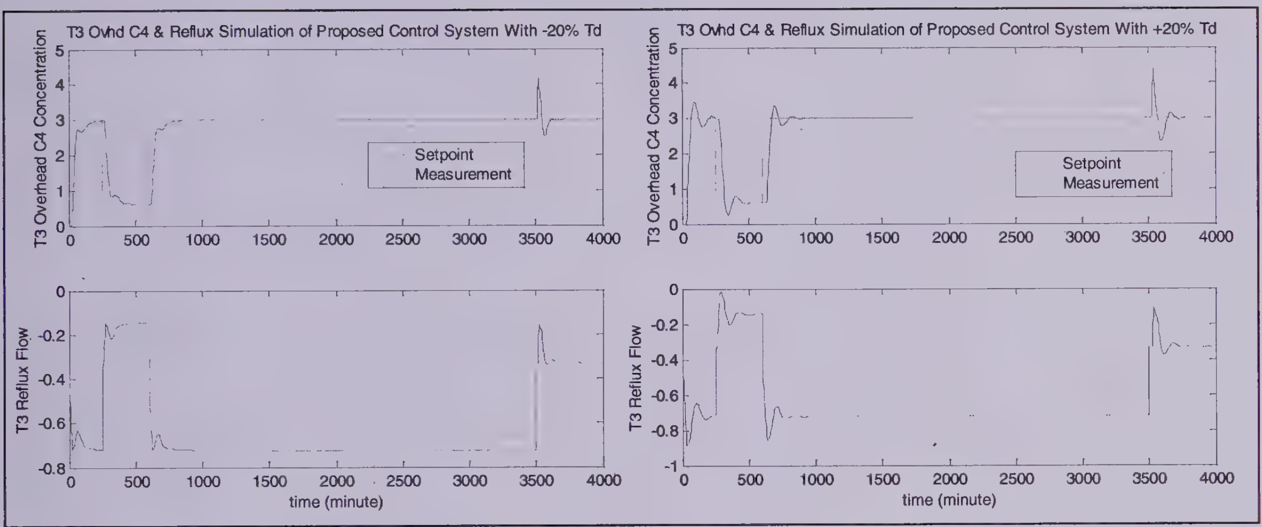

Figure 22. Comparison of T3 Ovhd C4/T3 Reflux Delay Time Mismatch

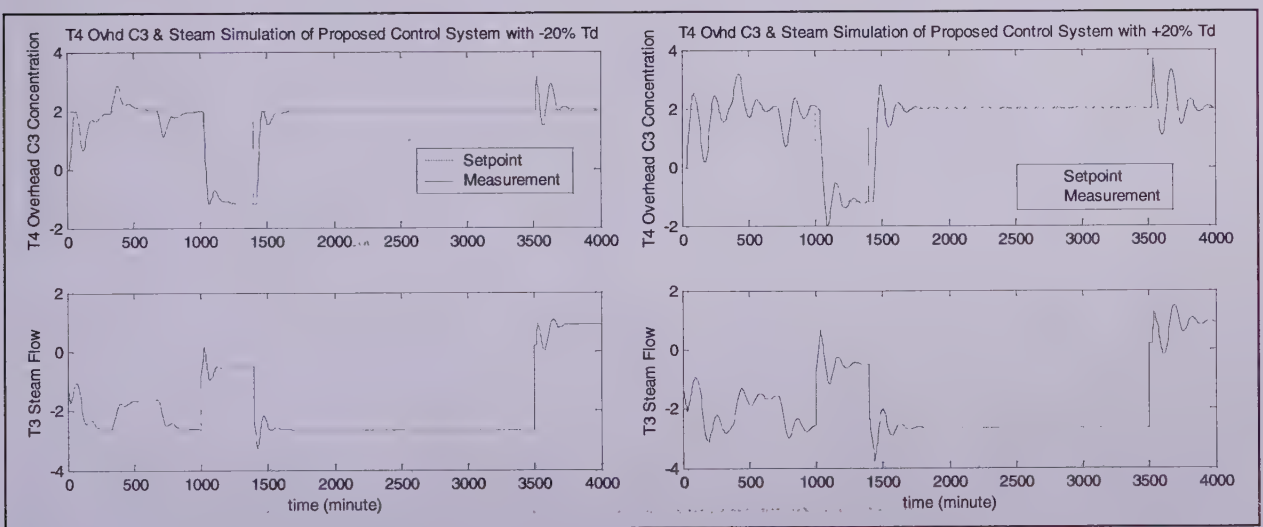

Figure 23. Comparison of T4 Ovhd C3/T3 Steam Delay Time Mismatch

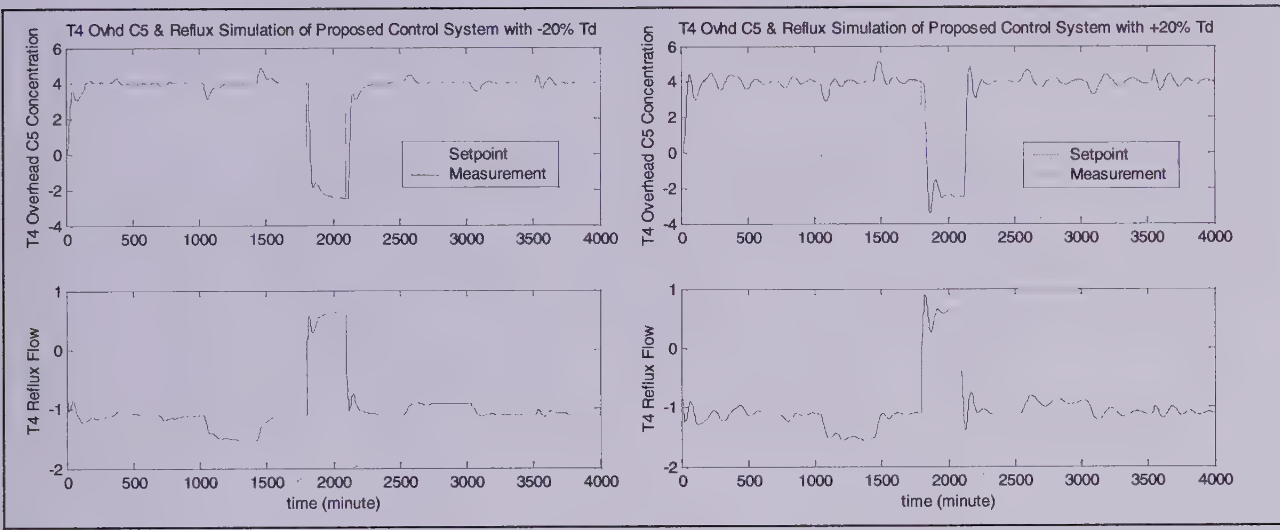

Figure 24. Comparison of T4 Ovhd C5/T4 Reflux Delay Time Mismatch

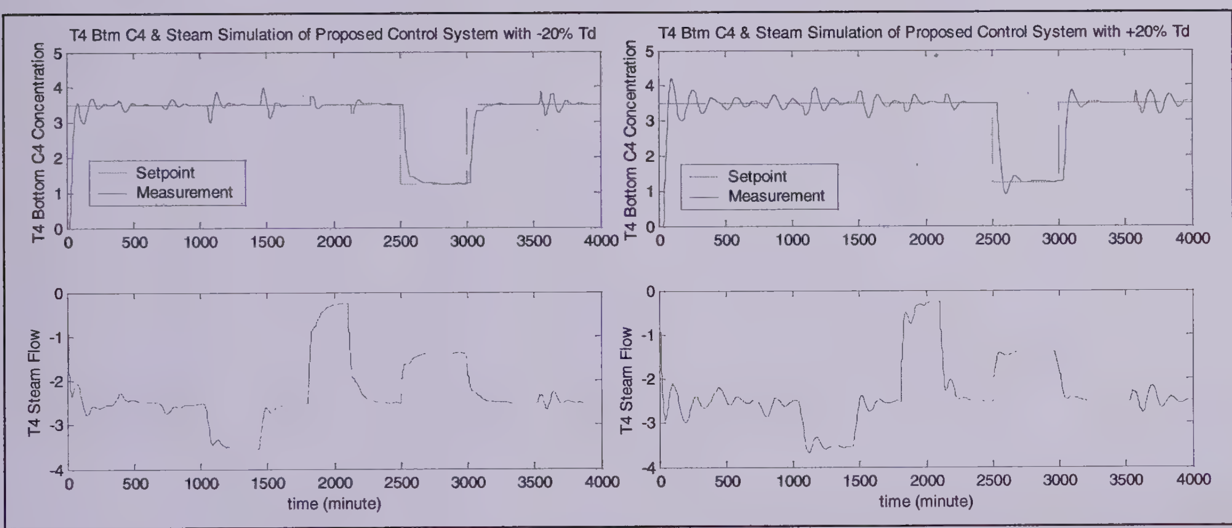

Figure 25. Comparison of T4 Btm C4/T4 Steam Delay Time Mismatch

As expected, a longer time delay would deteriorate the control performance since the deadtime between the disturbance and the process would also expand with the increments of delay time in all transfer functions. On the other hand, a 20% reduction of deadtime would shorten the delay difference between the two and allow the control feedback to respond faster. For the de-coupler in the T4 process, the mismatch of the deadtime for the reflux and steam loops would cause some degree of distraction to the process due to the untimely response of

the de-couplers, but the feedback control would compensate it. Thus, the impact was not significant in this case.

The reflux streams or the liquid-vapor traffic within the distillation towers could alter the process time delay by varying flow rates. With the same piping distance and diameter, an increase in flow would change the fluid velocity and thus reduce the deadtime in the process. Nevertheless, as found in this analysis, a shifting in the deadtime might affect the control performance slightly, but the feedback control would still perform quite well.

6. Results and Recommendation

In this study, it was first discovered that the sampling time for the Gas Chromatography analyzers time was quite long and might not be able to monitor the process appropriately. This deficiency could blindfold the process while waiting for the analyzer to update, and deteriorate control performance of the system. It is suggested to adopt the Inferential Control technique by applying the most responsive tray temperature of the towers to infer the product composition. This would provide sufficient process feedback on a continuous basis. Another potential enhancement is to implement the derivative action in addition to the PI controller in the inferential scheme. Since the temperature generally does not contain as much process noise as the flow or pressure, applying the derivative action could improve the control performance by delivering faster controller response.

The results from the RGA validated the control loop pairing in the existing control system. This was to use the reflux flow controlling the overhead composition and the steam flow manipulating the bottom purity of the tower. However, the RGA also revealed that there was significant loop interaction for T4 tower, especially in the T4 B C4 / T4 Steam loop. To counteract the undesired loop interaction, the de-coupler scheme could be implemented. Also, T3 C4 Feed and T4 O C3 were shown to have significant impacts to T3 and T4 respectively. Feedforward controllers were proposed to offset the negative effect from these disturbances. Nevertheless, steady-state feedforward controllers were implemented in some loops due to the larger deadtime or higher order of the process transfer function relative to the disturbance transfer functions.

With the implementation of the feedforward controllers and de-couplers in the system, the new control system was able to reduce the variability in the process and thus reduce the potential to produce off-specification products from the towers. In the simulation of the existing PI control system, when there is substantial C4 feed increment entering to T3, it would disturb T3 tower and raise the C3 composition higher than the quality specification. The upset condition would also pass on to the second tower causing the C5 component in T4 overhead higher than the quality constraint, while the bottom stream in T4 was barely able to meet the product specification. On the other hand, with the new control system, all control loops would be able to function effectively and allow the product streams for the towers to

meet the quality specifications comfortably. This could also reduce the process variability and open the door for further process optimization. Specifically, one might consider raising the control setpoints of some minor componenet compositions closer to their specification limits (i.e. move the setpoints toward the upper limit of 5). Since separation processes are always a tradeoff between the energy consumption and product purity, making less pure product would reduce the operating cost and improve the process efficiency. In practice, the setpoints for the composition control often depends on the product specification and performance of the controllers. If product specification is at 5% and the controller is able to maintain the concentration within 1%, the setpoint will often set at 4%. With the improved performance of the new control scheme, it is expected the variability of the product concentration will be reduced significantly, and thus, the existing setpoint can be raised toward their specification limits and the operating cost of the plant will be reduced. For a process that is capacity limited by the reboiler or condenser duty, this can even increase the process throughput without making any capital investment since the reduction in the steam or reflux consumption will make room for the plant to process more feed.

Based on the simulation results, a list of revised setpoints with each including a 0.5 safety margin is proposed in Table 4. Note that the suggested setpoint for T4_B_C4 is very close to the product quality limit. This is because the product stream exiting the bottom of T4 tower is not shipped to the customers directly, but is collected in a large storage tank, where the product can settle and mix, and thus does not impose any quality concern for a momentary process upset.

Table 5. Comparison of Tower Composition Setpoints

Controlled Variable	Existing Setpoint	Proposed Setpoint
T3_O_C4	3	3.5
T4_O_C3	2	3
T4_O_C5	4	4
T4_B_C4	3.5	4.8

Although the steady-state feedforward provided some degree of improvement on the control performance, there could be opportunities for further enhancement. One may explore the possibility to change the disturbance variable to others that can provide a feedforward controller capable of compensating the dynamics of the disturbance. This is to use any

inferring signals in the upstream so that the disturbance deadtime would become longer than the process deadtime. As discussed in the study, the T3 C4 feed stream, as a disturbance variable, had a process delay timer shorter than the process. Thus, the reflux or steam were not able to offset completely the disturbance before it reached the process. However, if there was another way to detect the feed change earlier in the upstream process, this would increase the delay time to the process and make the feedforward response more effective.

As demonstrated in Section 5 of this report, a model mismatch would influence the control performance. In general, a process gain mismatch could cause higher process variability on the system than the process time and delay mismatches. Yet, the proposed control scheme was able to compensate for the model discrepancy quite well. Anther point was that the model mismatch could also deteriorate the tuning of the controllers. Unless one can always maintain the tuning at the optimal setting, the tuning parameters should be kept at the setting that would preserve the plant stability at all times.

One should be aware that this study only focused on controlled and manipulated variables, and did not take other process constraints in the system into consideration. For example, the simulation would allow T4 steam flow to increase abruptly in order to bring T4 bottom C4 concentration down as fast as possible, but in reality, this could potentially create large vapor flow exceeding the hydraulic limit of the structure and lifting the internal trays. Another example is that the reflux, steam, and feed flows can be increased to their capacity limits in the simulation, but this did not consider the possibility of flooding due to excess tower loading. In the actual process, it is possible to see that the separation efficiency of the towers deteriorates significantly due to severe flooding and make the control system ineffective.

Finally, the simulation of the plant also verified that the capacities for the existing reboilers and condensers are sufficient. This was concluded from the fact that the steam and reflux were always within the established operating range throughout the simulation.

7. Concluding Remarks

Some opportunities were identified that could help to improve the control performance of the coupled distillation process. These include applying the inferential control technique to overcome the deficiency associated with the long analyzer sampling time, and implementing the feedforward and de-coupler scheme to compensate the disturbance and loop interaction respectively. These would lower the process variability and reduce the potential to generate off-spec products even when encountering a worse case of feed disturbance. In addition, the new control scheme would allow the process to operate more efficiently by raising the concentration setpoints of those minor components closer to their specification. This will lower the product purity slightly (still meeting product quality requirements), but at the same time reducing the operating cost of process with less energy consumption.

Although, it was known that the plant dynamic could vary frequently due to changes in operating conditions and could affect the effectiveness of the control system, the model mismatch study showed that the proposed scheme could still handle the offset with satisfactory results. It is also recommended that one should keep the tuning parameters for the PI controllers at the setting that would ensure stability at all times. This would prevent the process from becoming oscillatory due to a sudden change in feed condition.

8. NOMENCLATURE

C3+	Hydrocarbon liquid mixture containing propane and heavier components.
C4	Butane
C5	Pentane
CV	Controlled Variable is the variable containing the information about the internal state of the process that we want to change.
Kc	Proportional Gain is the gain of the proportional action in a PI controller.
Kp	Process Gain is the gain of the open-loop process system.
MV	Manipulated Variable is the variable that we can operate freely to stimulate the system and can induce change in the internal conditions of the process.
PI Control	Proportional Integral Control is a control algorithm that consists of proportional and integral actions. The proportional action responds proportionally to the offset between the setpoint and process output, while the integral action acts to the accumulation of the offset.
T3	Depropanizer Tower, a distillation tower, that receives the propane and heavier hydrocarbon mixture and separates it into two streams, a propane rich stream from the overhead and the heavy component stream from the bottom.
T4	Debutanizer Tower, a distillation tower, that receives the butane and heavier hydrocarbon mixture and separates it into two streams, a butane rich stream from the overhead and the heavy component stream from the bottom.
T_{CL}	Time constant of the closed loop process system.
Td	Time delay of the process system.
Ti	Integral time of the integral action in a PI controller.
Tp	Time constant of the open-loop process system.

9. REFERENCES

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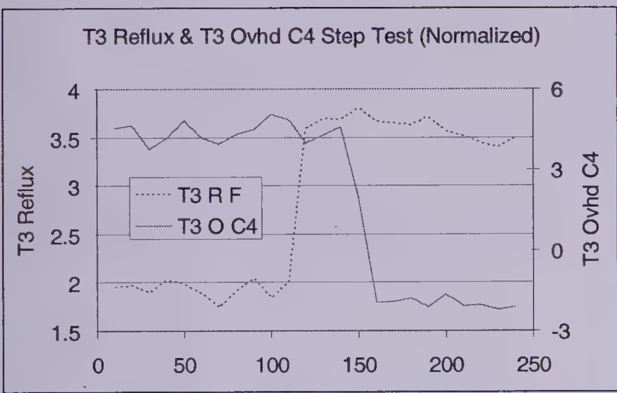
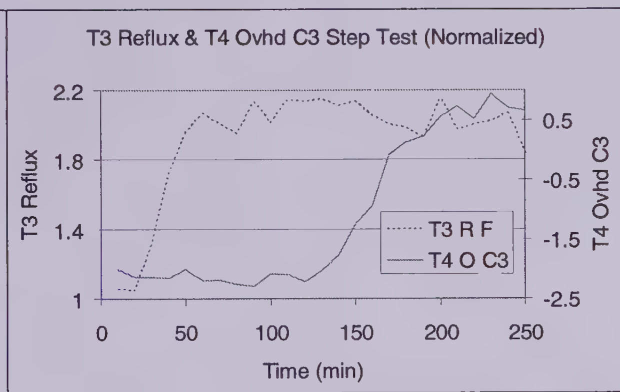
2. Raja Amirthalingam, Su Whan Sung, and Jay H. Lee, "Two-Step Procedure for Data-Based Modeling for Inferential Control Applications", *AIChE Journal*, **46**, 1974-1988, (2000)
3. K. L. Levien and Manfred Morari, "Internal Model Control of Coupled Distillation Columns", *AIChE Journal*, **33**, 83-98, (1987)
4. S. Kumar, J. D. Wright and P. A. Taylor, "Modelling and Dynamics of an Extractive Distillation Column", *The Canadian Journal of Chemical Engineering*, **62**, 780-789, (1984)

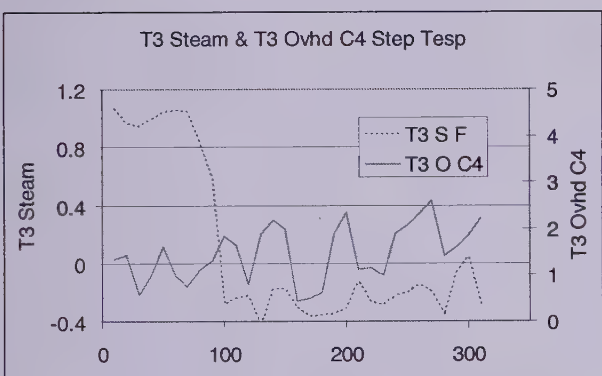
Appendix I. Graphs of Step Test Data

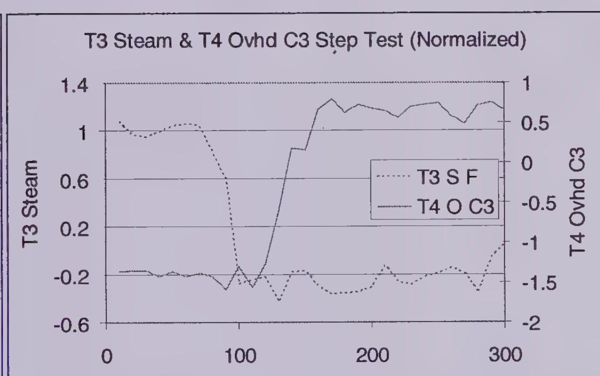
T3 Reflux Loops:



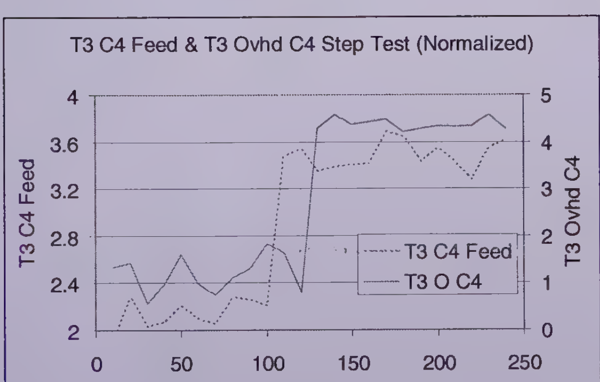


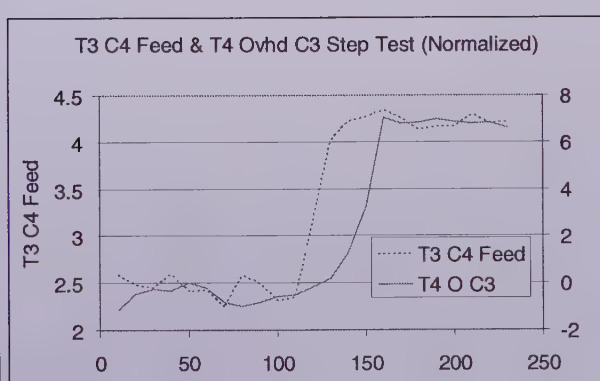
T3 Steam Loops:



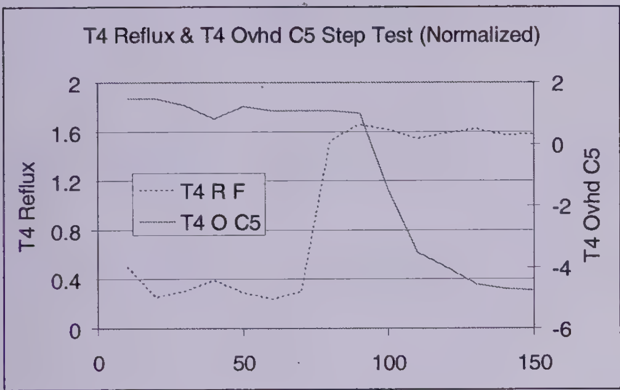


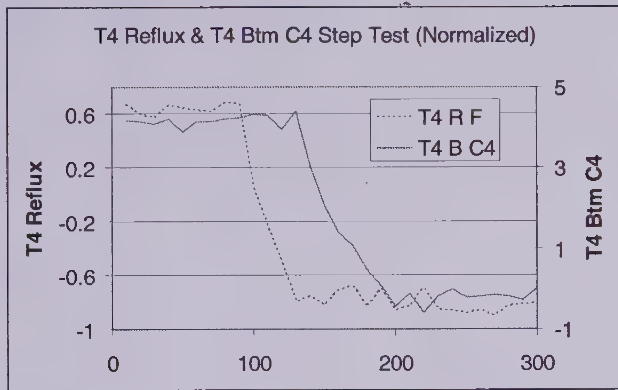
T3 Disturbance Loops:



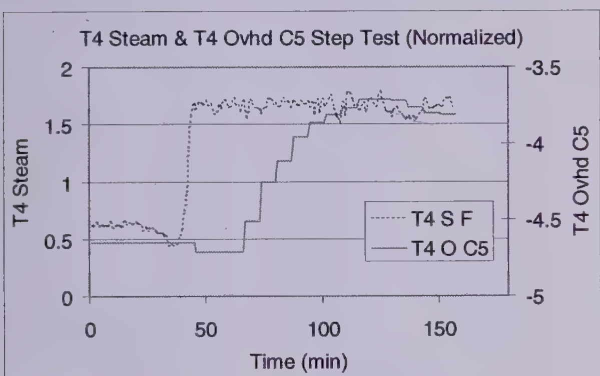


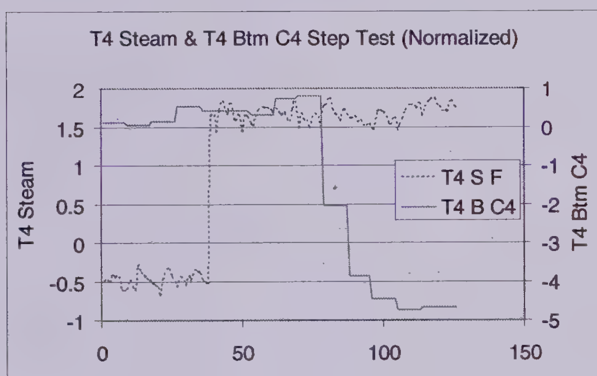
T4 Reflux Loops:



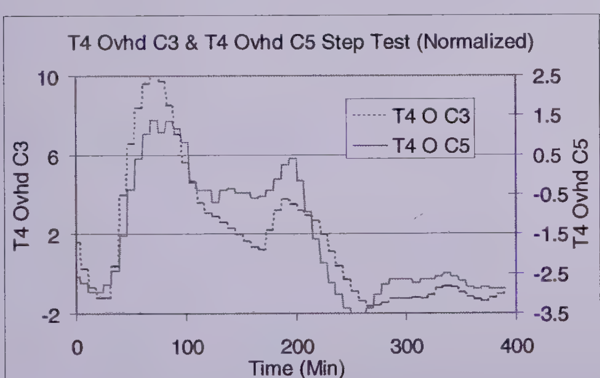


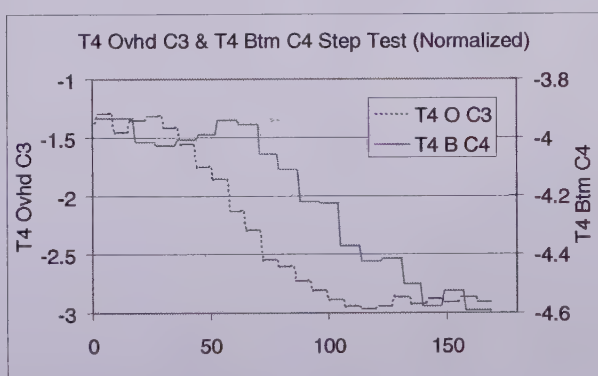
T4 Steam Loops:





T4 Disturbance Loops:





Appendix II. Relative Gain Array Analysis

Let Y be the steady-state gain of the 2 x 2 system: $Y = \begin{bmatrix} k_{11} & k_{12} \\ k_{21} & k_{22} \end{bmatrix}$

The RGA, Λ, for the 2 x 2 system, Y, can be calculated as:

$$\Lambda = \begin{bmatrix} \lambda_{11} & \lambda_{12} \\ \lambda_{21} & \lambda_{22} \end{bmatrix} \quad \text{where } \lambda_{11} = \lambda_{22} = \frac{1}{1 - \dfrac{k_{12}k_{21}}{k_{11}k_{22}}} \quad \text{and} \quad \lambda_{12} = \lambda_{21} = \frac{1}{1 - \dfrac{k_{22}k_{11}}{k_{21}k_{12}}}$$

Similarly, take the steady-state gain of the 2 x 2 system for each of the tower:

$$Y_{T3} = \begin{bmatrix} T3\ Ovhd\ C4 \\ T4\ Ovhd\ C3 \end{bmatrix} = \begin{bmatrix} -4.149 & 0 \\ 2.678 & -1.488 \end{bmatrix} \begin{bmatrix} reflux \\ steam \end{bmatrix}$$

$$Y_{T4} = \begin{bmatrix} T4\ Ovhd\ C5 \\ T4\ Btm\ C4 \end{bmatrix} = \begin{bmatrix} -4.76 & 3.288 \\ 0.798 & -2.531 \end{bmatrix} \begin{bmatrix} reflux \\ steam \end{bmatrix}$$

Therefore, by substituting the k element into the equation for λ, we can obtain the following:

For T3: $$\lambda_{11} = \lambda_{22} = \frac{1}{1 - \dfrac{0 * 2.678}{-4.149 * -1.488}} = 1 \quad \text{and} \quad \lambda_{12} = \lambda_{21} = \frac{1}{1 - \dfrac{-1.488 * -4.149}{2.678 * 0}} = 0$$

$$\Lambda_{T3} = \begin{bmatrix} 1 & 0 \\ 0 & 1 \end{bmatrix}$$

For T4: $$\lambda_{11} = \lambda_{22} = \frac{1}{1 - \dfrac{3.288 * 0.798}{-4.76 * -2.531}} = 1.278$$

$$\text{and} \quad \lambda_{12} = \lambda_{21} = \frac{1}{1 - \dfrac{-2.531 * -4.76}{0.798 * 3.288}} = -0.278$$

$$\Lambda_{T4} = \begin{bmatrix} 1.278 & -0.278 \\ -0.278 & 1.278 \end{bmatrix}$$

Appendix III. Feedforward / De-Coupler Design

1. The feedforward controller for $Gff_{31} = -\dfrac{T3\ Ovhd\ C4}{T3\ C4\ Feed}$ is determined as:

$$Gff_{31} = -\frac{Disturbance\ Transfer\ Function}{\Pr ocess\ Transfer\ Function} = -\frac{\frac{0.2}{6S+1}e^{-20s}}{\frac{-4.149}{13s+1}e^{-30s}}$$

However, the deadtime terms would give a net value of e^{+10s}, which is not valid, therefore a steady-state feedforward is used as alternative:

$$Gff_{31} = -\frac{0.2}{-4.149} = 0.048$$

2. The feedforward controller for $Gff_{32} = -\dfrac{T4\ Ovhd\ C3}{T3\ C4\ Feed}$ is determined as:

$$Gff_{32} = -\frac{Disturbance\ Transfer\ Function}{\Pr ocess\ Transfer\ Function} = -\frac{\frac{0.52}{23S+1}e^{-20s}}{\frac{-1.488}{21s+1}e^{-30s}}$$

However, the deadtime terms would give a net value of e^{+10s}, which is not valid, therefore a steady-state feedforward is used as alternative:

$$Gff_{32} = -\frac{0.52}{-1.488} = 0.349$$

3. The feedforward controller for $Gff_{41} = -\dfrac{T4\ Btm\ C4}{T4\ Ovhd\ C3}$ is determined as:

$$Gff_{41} = -\frac{Disturbance\ Transfer\ Function}{\Pr ocess\ Transfer\ Function} = -\frac{0.381e^{-40s}}{\frac{-2.531}{14s+1}e^{-30s}} = \frac{2.115s+0.15}{1}e^{-10s}$$

However, the order of the numerator is higher than the denominator term, which is not valid, therefore a steady-state feedforward is used as alternative:

$$Gff_{41} = -\frac{0.381e^{-40s}}{-2.531e^{-30s}} = 0.15e^{-10s}$$

4. The feedforward controller for $Gff_{42} = -\dfrac{T4\ Btm\ C4}{T4\ \mathrm{Re}\, flux}$ is determined as:

$$Gff_{42} = -\frac{Interacted\ Transfer\ Function}{\mathrm{Pr}\,ocess\ Transfer\ Function} = -\frac{\dfrac{3.288}{30s+1}e^{-40s}}{\dfrac{-2.531}{14s+1}e^{-30s}} = \frac{18.19s+1.3}{30s+1}e^{-10s}$$